MINIMIZATION OF ENERGY CONSUMPTION IN A CRYOGENIC PLANT THROUGH PROCESS MODIFICATION AND OPTIMIZATION USING A STEADY STATE PROCESS SIMULATOR

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INTRODUCTION

Over the last decade, pinch technology has been widely accepted as the principal tool in the integration of process plant for maximum energy recovery. However, due to the complexity of various processes and the interaction between them, the maximum benefit supposedly gained from this technology is often misquoted. An easier approach of optimizing energy requirement for a process plant is through modification of process operating conditions with respect to energy consumption and its integration with the background process.

Due to the growing importance of gas processing plant (GPP) in Malaysia, this paper takes an example of a local gas processing plant which produces pure natural gas from ethane rich natural gas feed stock. The heart of this GPP process is the low temperature separation units which separate the natural gas feed stock into its pure components at cryogenic condition. These energy intensive units consist of two heat exchangers, a cooler and a distillation column. A simplified schematic of the plant is shown in Figure1.

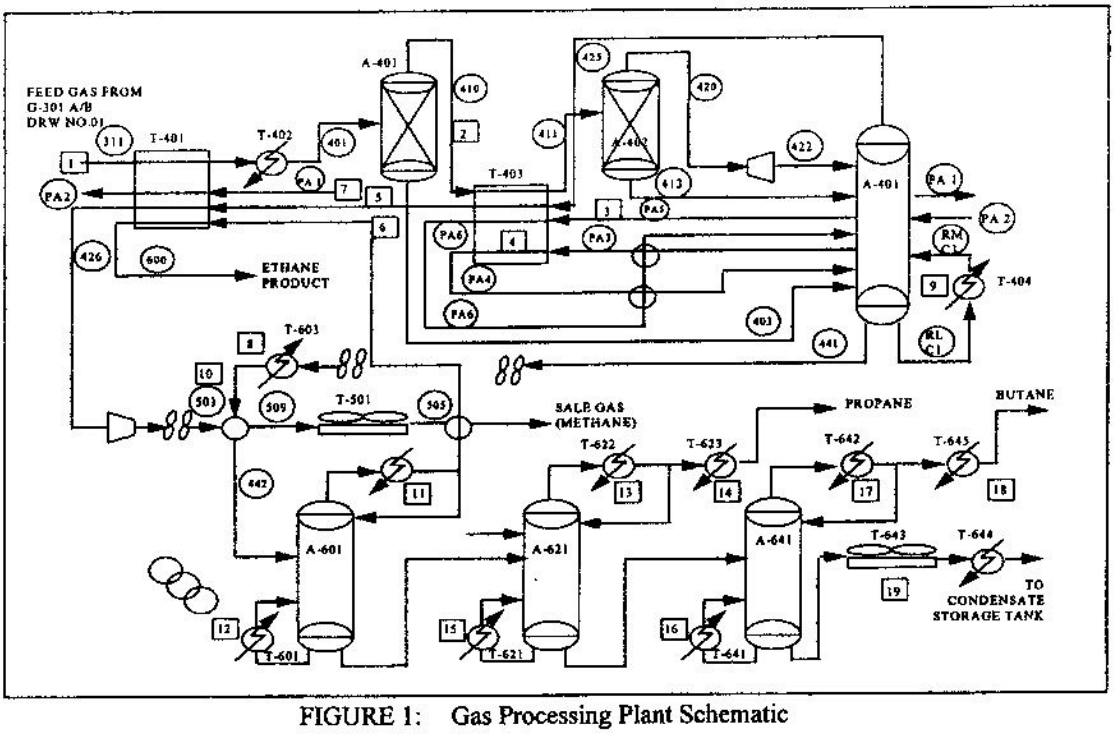

FIGURE 1: Gas Processing Plant Schematic

The objective of this work is to optimize the energy savings from the low temperature separation units by:

i) reducing the refrigerant duty in the cooler by exchanging a maximum amount of heat in both heat exchangers ; and
ii) modifying the distillation column operating conditions without altering the final product specifications.

The application of process simulation software in this work has effectively guided the determination of optimum process conditions in order to achieve the above objective.

METHODOLOGY

The low temperature separation units were modified and optimized using a steady state process simulator with respect to energy. Figure 2 represents the process units involved in these modifications. Our target is to reduce the refrigerant load of a cooler T-402 without altering the product specifications of demethanizer column A-401.

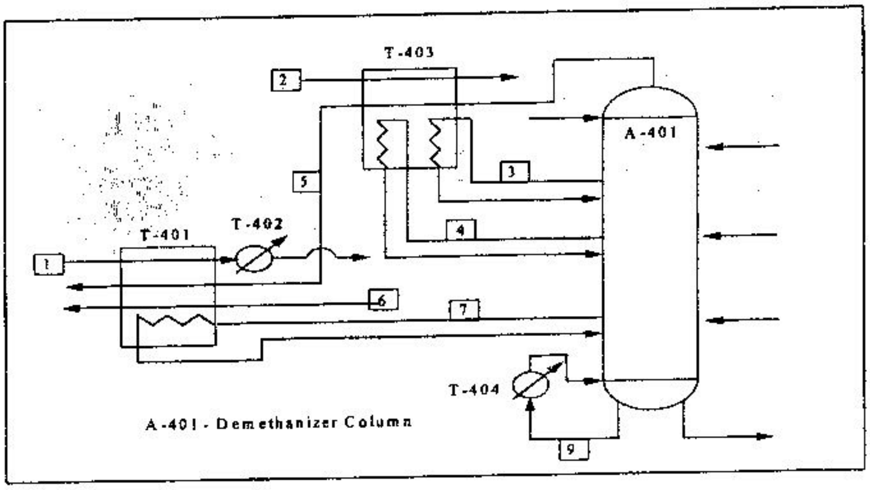

FIGURE 2: Scope for Process Modifications

The above process flow diagram clearly depicts the direct relationship between the duty of cooler T-402 and the target temperature of stream 1. Higher stream temperature would result in greater refrigerant load of T-402. Streams 5, 6 and 7 which are exchanging heat with stream 1, also have a direct impact on T-402. If we can make these streams absorb more heat from stream 1, its target temperature will be reduced and so does the cooling requirement of T-402. However, we found that only streams 5 and 7 have potentials for modifications. Stream 6 turns out to be the top product stream of deethanizer column and therefore could not be modified. Figure 3 demonstrates the effect of these modifications on units T-401 and T-402

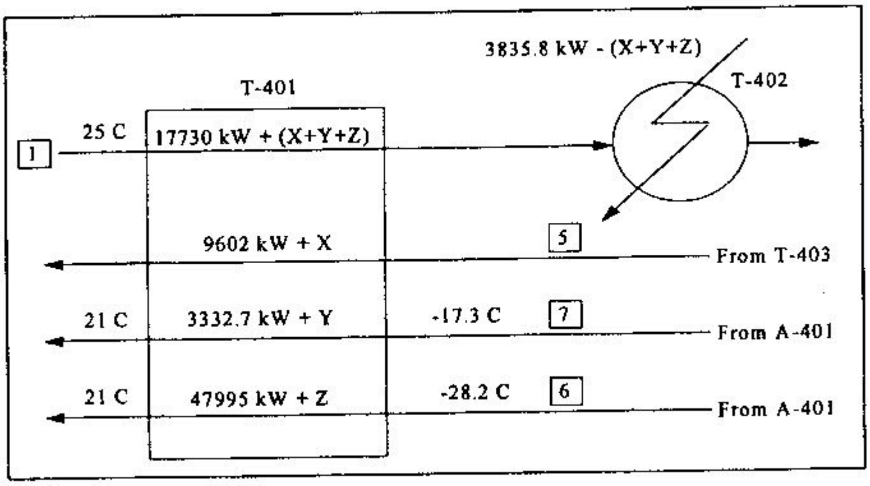

FIGURE 3: Effects of Stream Modifications on Units T-401and T-402

Further inspection also reveals that stream 5 is more sensitive to changes than stream 7 since it is stretched across two heat exchangers. Any modification made on stream 5 heating load at unit T-401 has to be balanced with its load at T-403. This scenario is presented in Figure 4. Streams 3, 4 and 7 are the pump-around streams from demethanizer column, A-401. Changes on the conditions of these streams will affect the operation of the column and hence the product specifications.

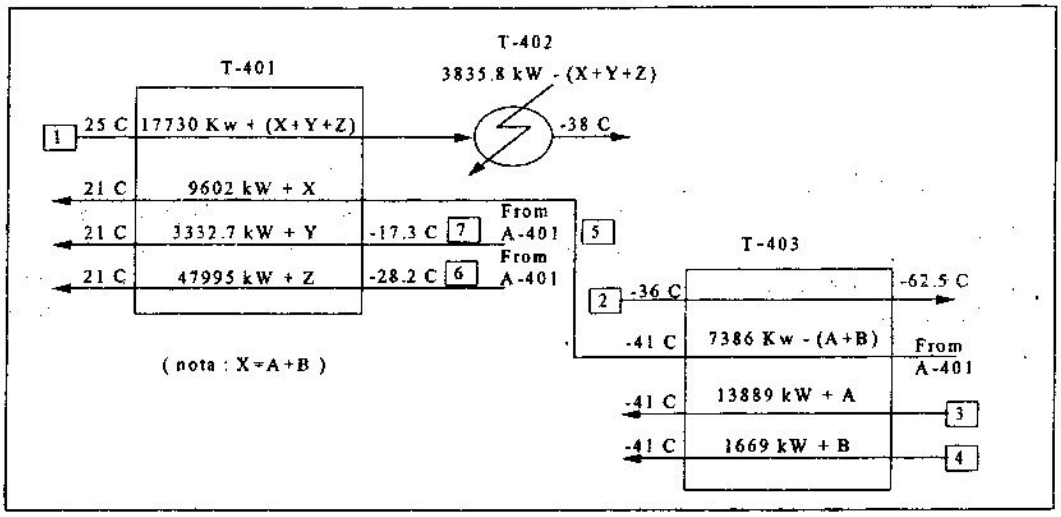

FIGURE 4 : Effects of Stream Modifications on Unit T-403

PROCESS SIMULATION

In order to maintain the final product specifications of demethanizer column, process simulation has to be carried out for the entire low temparature separation units to determine the effects of column modifications on other units. The use of simulation program enables us to consider different alternatives in the search of optimum column operating conditions. The input data for the simulation runs are taken from the company's process flow diagram (PFD) and is given in Table 1.

TABLE1 : Summary of Distillation Column Stimulation Input Data

Items	Base Case	Optimum Case
Thermodynamic option set	- Peng-Robinson (PR)	- Peng-Robinson (PR)
Feed stream data	- Flowrate,Temperature,Pressure and components fraction	- Flowrate ,Temperature, Pressure and components fraction
Demethanizer Column	- Numbers of trays (41 trays) - Reboiler Duty (13.8088 MM kJ/h) - Side Stream Flowrate (stream 3,4 &7)	- Numbers of trays (41 trays) -Top Product Flowrate (13243 kmol/h) - Side Stream Flowrate (stream 3,4 & 7)
Heat Exchanger , T-401	- Targeted Temperature of stream 5 & stream 6 (21 oC) - Stream 7 Duty (11.9977 MM kJ/h)	- Targeted Temperature of stream 5 , 6 & 7 (21 oC)
Heat Exchanger, T-403	- Targeted Temperature of stream 2 & stream 4 (-62.5 oC , -38oC) - Stream 3 Duty (5.0002 MM kJ/h)	- Targeted Temperature of stream 2, & stream 3 & stream 4 (-62.5 oC , - 41oC - 41oC)
Cooler , T-402	- Stream 1 targeted temperature (-36oC)	- Stream 1 targeted temperature (-36oC)

RESULTS AND DISCUSSIONS

The simulation results are tabulated in Table 2. As mentioned earlier, our simulation objectives are to maintain the product specifications at the top and bottom streams of column A-401 as well as the target temperatures of streams 1, 5, and 6. Other stream conditions therefore, would be inconsistent with the original data. The differences between the original data and the optimum conditions are highlighted in Table 3. The reductions in distillation column reboiler duty and refrigerant duty of unit T-402 are calculated to be 929 kW and 1356 kW respectively. The savings which are attributable to lower fuel and refrigerant costs amount to approximately US$776000 per annum. There is no additional capital outlay at stake in this savings.

TABLE 2 : Distillation Column Simulation Summary

Variables	Material and Energy Balance	Base Case (without modification)	Optimum Case
- Top Product Flowrate (kmol/h)	13243.54	13193.82	13242.91
- Top Product Composition(methane mol fraction)	0.973963	0.97428	0.97422
- Top Product Methane Flowrate (kmol/h)	128 98.72	12854.34	12901.51
- Top Product Stream Temperature (oC)	-92.9402	-93.0740	-93.0476
- Bottom Product Flowrate (kmol/h)	2889.78	2939.53	2890.20
- Bottom Product Stream Temperature (oC)	22.8789	17.7361	23.1617
- Reboiler Duty (kW)	3835.78	3835.78	2908.22
- Stream 3 Duty (kW)	1388.94	1388.94	1589.97
- Stream 4 Duty (kW)	1666.89	1250.39	949.86
- Stream 7 Duty (kW)	3332.69	3332.69	4782.64
- Unit T-402 Duty (kW)	3680.00	4021.11	2664.72
- Stream 1 Target Temperature (oC)	-36.0	-36.0	-36.0
- Stream 6 Target Temperature (oC)	21.0	21.0	21.0
- Stream 5 Target Temperature (oC)	21.0	21.0	21.0

TABLE 3 : Simulation Results Before anf After Column Modifications

Variables	Base Case	Optimum Case	Differences	% Differences
Top product flowrate (kmol/h)	13193.82	13242.91	49.09	+0.37
Bottom product flowrate (kmol/h)	2939.53	2890.20	49.33	-1.68
Top product concentration (CH_4 mol fraction)	0.97428	0.97422	0.00006	~ 0
Top product CH_4 flowrate (kmol/h)	12854.34	12901.51	47.17	+0.37
Reboiler Duty (kW)	3835.78	2908.22	928.56	24.21
Unit T-402 Duty (kW)	4021.11	2664.72	1356.39	33.73
Stream 3 Duty (kW)	1388.94	1586.97	198.03	14.26
Stream 4 Duty (kW)	1250.39	949.86	300.53	24.03
Stream 7 Duty (kW)	3332.69	4782.64	1449.95	43.51

CONCLUSION

Successful implementation of process modification and optimization with respect to energy consumption requires a careful analysis on the interactions between the process units involved. Our study on a local gas processing plant demonstrated a sensible proof of the value of process integration technology to the industry. The annual savings generated from simple process modifications on GPP process are estimated over US$776000 with no investment nor risk involved. The simulation program has provided greater opportunities for flexible process operation of which could be transformed into substantial savings to the company.

REFERENCES

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